

Liquid Transport on a Nylon Surface Moving Between Liquids Through an Air-filled Chamber

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The flow rate of liquid carried on a nylon strand moving from a liquid filled space, through an air filled space, to another liquid filled space was determined from measurements of electrical conductance of the separated liquids. The measurements are consistent with the transmission of a liquid layer on the nylon surface of constant thickness, independent of surface velocity. The layer thickness is low enough that an adsorptive strand can carry adsorbate from a solution at a practical rate. The effective concentration of solution taken from an adsorption chamber by a saturated adsorbent strand is estimated to be at least $5.8 (\lambda/S)$ times the chamber concentration, where λ is the chamber length and S the line speed.

Le débit de liquide se déplaçant sur une fibre de nylon depuis un espace rempli de liquide vers un espace rempli d'un autre liquide, à travers un espace rempli d'air, a été déterminé d'après les mesures de conductance électriques des liquides séparés. Les mesures concordent avec la transmission d'une couche liquide à la surface du nylon d'épaisseur constante, indépendamment de la vitesse de surface. L'épaisseur de la couche est assez mince pour qu'une fibre adsorbante puisse transporter de l'adsorbat depuis une solution à un taux acceptable. On estime que la concentration effective de solution extraite d'une chambre d'adsorption par une fibre adsorbante saturée est d'au moins $5,8 (\lambda/S)$ fois la concentration de la chambre, où λ est la longueur de la chambre et S la vitesse continue.

Selective methods of mixed solute separation based on adsorption are available on laboratory scales, usually based on fixed beds of porous adsorbents of increasingly sophisticated structure (Pfeiffer, et al., 1996), but more recently including fibrils (Kobuke, et al., 1988) and membranes (Thommes and Kula, 1995). These methods are too expensive to separate large amounts of product, although preparative scale columns are available to separate sufficient quantities of some pharmaceutical compounds or mixtures to satisfy commercial requirements. The rate limitation in these methods is the liquid-solid contact rate, that is, the time for the solute to diffuse through a stagnant liquid in the pore. If nonporous particles are used the particles can be made smaller in diameter to increase surface area, but if size is reduced to that required to make their capacity comparable to porous particles, flow resistance increases exponentially and the fraction of liquid within the bed which is stagnant increases. Although membrane systems allow a more consistent pore size and spacing to be achieved, which minimizes flow resistance and isolated liquid spaces for a selected pore size and number, the membranes still exhibit the surface area versus flow rate tradeoff, that is, increasing flow restriction as area increases.

A significant consideration for the solute recovery process is the capacity of the adsorbent to absorb solute. As the sorbent nears saturation the flows must be switched and the solute desorbed and transferred to a product stream. This requires either a cyclic batch or a countercurrent process, with the adsorbent moving out of the adsorption vessel to a regeneration process. Countercurrent operation is preferable

for commercial scale processes because the equipment is used continuously. The countercurrent contact can be simulated by moving the feed and draw-off points along a fixed bed. The simulating bed is usually divided into a substantial number of sections, so the simulated motion of the breakthrough front is discrete (Saska, et al., 1992). Interstitial flow rates of fluid through the bed of up to 6 mm/sec. have been reported (Ching, et al., 1985) for this type of process.

The flow resistance resulting from the particle/liquid contact can be reduced by suspending the particles using countercurrent contacting equipment, such as a rotary disc contactor (RDC). In RDCs the heavier solid is fluidized by the lighter liquid phase and moves countercurrently downward, agitated by rotating discs mounted on a central axis. As with packed beds, mass transfer rates depend on interparticle diffusion of the solute components and thus on particle porosity and size. Similar to packed beds, optimum particle size is a compromise. Reduced size increases surface area and mass transfer rate but reduces settling velocity of the particles in the liquid (which controls flow rates of liquid and particles through the RDC). In a test series using hexane and gel catalysts the maximum liquid velocity permitting downward flow of the particles in the size range of 0.30–0.42 mm was measured to be around 2 cm/sec (Blytas, 1988).

High rates of contact can also be achieved by moving the solid rather than the liquid. A process and device for promoting adsorption by continuously contacting a solution with a moving adsorptive strand was described (Dickey, et al., 1988), the strand carried adsorbate through a low friction air lock and into a liquid filled desorption chamber. The feed solution was pumped through a vessel or chamber through which the adsorptive strands moved; strand motion was independent of the liquid flow rate and adsorbed material was desorbed in a comparable but separate chamber held at conditions promoting desorption. An advantage of this process over moving adsorbent particle methods is that fragile adsorbent surfaces, on the strand, can be used since

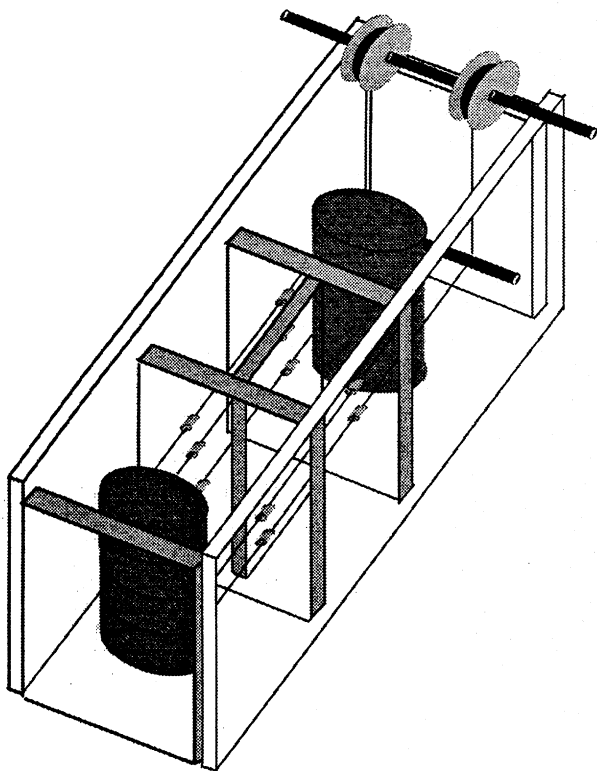


Figure 1 — Apparatus for moving strand through liquid filled chamber with minimal liquid flow between chambers: line drive mechanism and chamber layout.

there is no sliding contact or collision of the surface with other solids. To avoid abrasion of the continuous solid strand surface, the air sealing method must be effective in minimizing liquid transmission between liquid chambers. This article describes a simple test which provides an estimate of the rate of liquid carried by a moving strand in a laboratory scale adsorptive device. Gas sealing is emerging as the preferred method of sealing rotating equipment where maintaining high product purity is important (Adams, 1996); more elegant and effective designs useable for adsorptive separation are to be expected.

Experimental set-up for the determination of liquid entrainment

Experimental testing was carried out using the equipment shown in Figure 1, which shows only three loops for easier recognition. Commercial fishing line (Berkley "Trilene XL", 6 Kg, .036 cm diameter) comprised the adsorbent strand which was pulled from the delivery spool to the take up spool using a variable speed drive motor. The spools could be reversed (along with the direction of strand travel) by changing the one spool that was fixed to the drive shaft, using a set screw, and releasing the other. The line is coiled into seven horizontal and hemicircle ended loops which are shaped by vertical 47 cm diameter rollers on either end of the loops. The loops, spaced vertically about 0.7 cm apart, run through air chambers formed by the intersection of 0.07 cm diameter cylindrical holes drilled in 1.2 cm thick plexiglass blocks separating four adjoining liquid-filled chambers. Figure 2 is a drawing of half of one of the central chamber ends, shown in Figure 1. Air supply holes were drilled from the sides of the block. The chambers are fixed together with stainless steel screws which connect adjacent

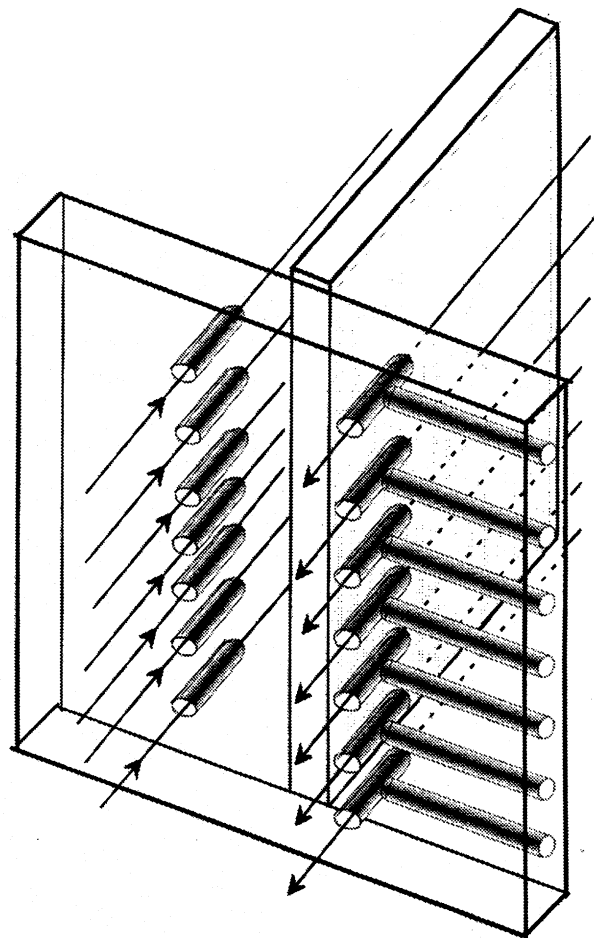


Figure 2 — End block of chamber showing air-sealed holes and intersecting air delivery ports (right hand side).

blocks through holes in rubber gaskets between the blocks (which comprise the vertical sides and the base); this design allows the structure to be disassembled in order to modify attachments, such as roller supports for aligning the strand feeds and returns to the spools which are held above one end of the structure. Compressed air can be pumped through the seven holes in one block at individually adjustable rates, controlled by separate valves and rotameters attached to the inlets of the holes.

MEASUREMENTS

Experiments were carried out by filling all four chambers to an equal height with distilled water. Air flow to the seven inlet ports was then adjusted so that air was just barely bubbling into the liquid in one adjacent chamber or the other — the air was introduced into the middle of each hole in the blocks and could emerge on either the right or left side. The initial bubbling flow rate was adjusted by trial and error because the seven ports were aligned vertically and the bubbles emanating from higher holes had a lower back pressure to overcome to initiate bubbling at the lower holes. Once the air flows to the holes was properly adjusted, a concentration of sodium chloride solution was added to one chamber to function as a flow tracer. The line drive motor was turned on and conductance readings in the chamber containing the tracer and adjacent chambers were taken at convenient time intervals. The measurements were taken with specific conductance electrodes placed close to the barrier between the

TABLE 1
Experimental Results for the Liquid Flow Rate Between Chambers

b (10^{-6} s^{-1})	S (10^{-2} m/s)	L ($10^{-9} \text{ m}^3/\text{s}$)	air flow	liquid viscosity $\text{Pa} \cdot \text{s}$
10.0	4	8.0	no	1000
5.7	-4	4.6	yes	1000
5.0	4	4.0	yes	1000
3.9	2	3.1	yes	1000
2.4	0	1.9	yes	1000
10.4	4	8.3	no	2000
10.0	4	8.0	yes	2000

high solute concentration chamber and the chamber downstream. The electrodes were moved briskly back and forth to stir the liquid in the chamber and to clear the electrode. This allowed us to obtain reproducible readings.

Since the same amount of liquid is carried into and out of the chamber by the same process at either end of each chamber, the mass in each chamber remains constant, and is unchanged by the passage of the line. The low salt concentration has a negligible effect on liquid density. Two tests were made in which carboxymethyl cellulose (cmc) was added to the initial water charge to determine the effect of viscosity on the liquid transfer rate. The viscosity of the cmc solution was determined from time measurements of sample flow through a calibrated capillary tube.

Line speed, S , was determined by measuring the rotation rate of the takeup reel and multiplying it by the time averaged circumference of the take up reel hub and wrapped strand, during a typical 20 minute run. The effective diameter of the take up reel, which turned at a constant rotation rate, increased from a minimum of 3.1 cm. at an average rate of 0.036 ms^{-1} .

CALCULATIONS

Changes in the conductance of the chamber originally containing the salt were converted to a liquid transfer rate based on a model of the chamber as a perfectly mixed, constant volume vessel with solute loss rate equal to the instantaneous liquid concentration divided by the chamber residence time, L/V :

$$\frac{d}{dt} C = C \frac{L}{V} \dots\dots\dots (1)$$

The time and conductance measurements were fit to the form:

$$y = a[\exp(-b \times t)] - c. \dots\dots\dots (2)$$

The conductance was linearly dependent on salt concentration for the concentrations used. The values of the exponential coefficient, b , are listed in Table 1. Although the experiments were batch processes, they were only run long enough for the variation in tracer concentration to closely simulate a steady-state process with concentrations about the same as the initial concentrations; typically a 20 minute run would produce an essentially linear drop in conductance of 3-5% (the form of equation 2 is nearly linear for low values of the exponent, $b \times t$). With conductance directly proportional to salt concentration the exponential coefficient b can be identified as the residence time of the salt in the chamber, and the liquid flow rate, L , determined from:

$$L = Vb = 0.8 \times 10^{-3} b \dots\dots\dots (3)$$

Experimental results

AIR FLOW RATE VARIATION

The air flow rate was not measured because good evidence was obtained in preliminary tests that the minimal air flow that could assure that a hole was filled with air is preferable to higher rates. Air flow rates above the minimum required rate tended, in some cases, to increase liquid transfer above the no air condition (bubbling direction can be unstable and shift from one chamber to the other carrying liquid with it - as long as it just bubbles on one side no liquid transfer due to the bubbling itself occurred). The minimal air pressure necessary to overcome capillary force should be the same for all of the ports with an additional backpressure provided by the hydrostatic head of liquid at the hole depth in the liquid.

LINE SPEED VARIATION

Values in Table 1 show the expected increase in L with line speed, S :

$$L = 1.9 + 0.5 \cdot S. \dots\dots\dots (4)$$

In the second test listed in Table 1 the direction of line motion was reversed, giving a slightly higher liquid flow rate, this might be due to some difference in the alignment of the line in the hole or effect on the bubbling.

VARIATION OF INTERFACIAL STRENGTH

The highest liquid flow rate values result when no air is used and when the liquid viscosity was raised. In these cases the fluid in the hole was less resistant to the pumping effect of the line moving through the hole.

Implication of liquid flow rates for adsorption

The process of removing solute from a body of liquid using the adsorptive strand method can be envisioned as two steps: (1) concentrating the solute in a thin film coating the strand and (2) carrying this film to the next chamber. The practicality of the process will depend on minimizing the bulk liquid carried along with the enriched adsorbed film layer, and adsorbing the solute on the strand at rates consistent with the adsorbent strand speed.

The linear relation between flow rate and line speed (for the range listed) is consistent with a model of the liquid flow comprised of a constant background rate, identified as the 0 line speed rate, and a rate resulting from a layer of liquid of constant thickness, m , carried on the line through the hole in the barrier. The layer thickness will depend on the viscosity and surface tension of the liquid. But for a particular liquid with its properties fixed:

$$L = L_0 + 4 S/\pi[m + d]^2 - d^2] \dots\dots\dots (5)$$

The thickness of the layer, m , in the tests reported here was calculated from the liquid flow rate (for 7 lines) and that line diameter to be $1.3 \times 10^{-5} \text{ m}$. For gravity draining films, especially those on plates or cylinders pulled vertically from a liquid pool, the literature has been recently summarized (Quere, et al., 1990; Brochard-Wyart and de Gennes, 1992).

One conclusion of the work described in these articles is that the film thickness should increase with the velocity of the moving solid, in proportion to $S^{2/3}$. The constancy of the hypothetical film in these experiments with horizontal moving lines may be due to the slight pressure resistance confronting the film as it enters the air filled space between chambers. With layer thickness independent of line speed, solute separation effectiveness will only depend on the hydrodynamics within the absorption chamber and the intrinsic rate of the adsorption at the surface. A measure of separation effect is the ratio: concentration of solution carried out (on the strand and including the film on the strand) to the concentration of solution fed to the adsorption chamber. For lack of particular adsorption data we assume the liquid layer on the strand is completely depleted of solute, which is the worst solute distribution situation from the standpoint of transferring solute as a concentrate to the next chamber. The mean concentration carried out in a steady process would equal the rate of adsorption divided by the liquid flow rate. For fixed beds the rate of solute moving to the surface is often calculated from one of a number of semi-empirical correlations that predict Sherwood number from Reynolds (Re) and Schmidt (Sc) numbers, which can usually be estimated. The hydrodynamic features of mass transfer to continuous, moving solids have been well studied (Sakiadis, 1961; Erickson, et al., 1966; Gorla, 1980; Adamczyk, et al., 1982). Gorla has provided numerical solutions for unsteady state mass transfer to a continuous moving sheet which are appropriate for high Schmidt number, aqueous liquid, cases. Hydrodynamic estimates indicate that much greater mass transfer rates can be achieved in continuous moving substrate devices than in conventional packed bed adsorbers with similar adsorption chemistry. The adsorption rate limited by intrinsic rates of the solute attaching to the solid will depend on the size, flexibility and other properties of the solute, and the adsorbing surface. Reports of direct measurement of intrinsic adsorption rates from solution are not plentiful, however estimates of a film coefficient, k_f , are made from models of porous, fixed bed adsorption data, where the rate is given by

$$q = k_f A C \dots \dots \dots (6)$$

A typical value of k_f (Olmstead and Weber, 1990) is 10^{-5} ms^{-1} , which, combined with the (7) strand area of $2.9 \times 10^{-4} \text{ m}^2$ per m of chamber length, for the chamber studied here, gives an adsorption rate of $2.9 \times 10^{-9} \lambda C \text{ kg/s}$. The carryover of liquid by the strands moving through the same chamber, based on the measurements of liquid flow, in Table 1, (with air flow to holes) is approximately $0.5 \times 10^{-9} S \text{ m}^3 \text{ s}^{-1}$, assuming that the 0 strand speed flow, due to leaks, could be eliminated with a tighter chamber design, reducing all listed flows by that amount. The concentration of solute on a strand saturated to the intrinsic limit carried to the next chamber is therefore, at least $5.8 \lambda/S$ times the feed solution concentration, C . The effectiveness can be seen to increase with chamber length (up to a limit of the intrinsic rate saturation), and by decreasing line speed, which is undesirable from a capital cost standpoint. The ratio λ/S is a residence time for the strand in the chamber, similar to the residence times used for modeling volumetric flows to vessels.

Actual tests with an adsorbing system are necessary to validate these estimates, based on a very simple model, which indicates the method could be used effectively to

remove solute from the solution in an adsorption chamber and concentrate it downstream in a parallel desorbing chamber. The complicated part of the equipment is the strand driving mechanism and the air-sealed hole containing end plates. The chambers can be lengthened, as necessary, using standard pipe with internal divider or two pipes, to bring the strand and liquid into equilibrium at the strand exit end. Therefore the cost effectiveness of the moving strand device will be less sensitive to the particular adsorption/desorption kinetics as the scale increases.

Nomenclature

A	= adsorbent surface (m^2)
a, b, c	= coefficients for numerical fit
C	= solute concentration (kg/m^3)
d	= line diameter (m)
d_p	= particle diameter (m)
D_L	= liquid diffusivity of solute (m^2/s)
k_f	= external film coefficient of liquid layer on line (m/s)
L	= liquid flow through holes in barrier (m^3/s)
L_0	= liquid flow when $S = 0$ (m^3/s)
m	= thickness of liquid layer on line (m)
q	= adsorption rate (kg/s)
S	= line speed (m/s)
t	= time (s)
V	= volume in chamber (m)
y	= conductance of solute rich chamber chamber – conductance of upstream chamber (mS)
z	= interstitial flow rate (m/s)

Greek symbols

λ	= length of a strand in a chamber (m)
ν	= kinematic viscosity (m^2/s)
σ	= surface tension (J/m^2)

Dimensionless numbers

Ca	= capillary number ($S \nu / \sigma$)
Re	= Reynolds number ($z d_p / \nu$)
Sc	= Schmidt number (ν / D_L)
Sh	= Sherwood number ($k_f d_p / D_L$)

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